B.Tech thesis report on

SIMULATION AND ECONOMIC ANALYSIS OF CRUDE DISTILLATION UNIT USING ASPEN PLUS

In partial fulfillment of the requirements for the degree of

Bachelor of Technology
in
Chemical Engineering

Submitted by

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(110CH0084)

Under the supervision of

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Department of ChemicalEngineering

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CERTIFICATE

This is to certify that the thesis entitled "**Simulation and Economic Analysis of Crude Distillation Unit using Aspen Plus**" submitted by **Mr. Prabin Kumar Pradhan (110CH0084)** in partialfulfillment of the requirements for the degree of Bachelor of Technology in Chemical Engineering at National Institute of Technology, Rourkela is an authentic work carried out by him under my supervision and guidance.

To the best of my knowledge, the matter embodied in this thesis has not been submitted to any other university or institute for the award of any degree.

Date:

Place: Rourkela

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ABSTRACT

Crude Distillation Unit (CDU) system is considered to be the most fundamental process of petroleum refining. This work aimed at the general thumb rules for the grass-root design of crude distillation unit (CDU) using aspen plus. Crude distillation unit system constituted a pre-flash tower, an atmospheric distillation tower and a vacuum distillation tower. Temperature and pressure profiles for all the three towers were considered to know the desired product tray. Here two crudes were used namely BOMBAYHG (Bombay crude), ARABY (Arabian Light Crude). Different binary combinations of these crude were used as feed to this crude distillation unit system. The True Boiling Point curves were obtained for each binary feed were obtained to access the variation in the compositions of the feed used. For simulation various product design specifications like for naphtha, HNAPTHA, diesel were considered based on the ASTM regulations and target temperatures for these products were also considered. After simulation, the objective function considered was the profit function, the prices of all raw materials, the steam used, the power cost and the income from the products were considered to calculate the profit. This profit was calculated with flow constraints on steam used in the towers. This work provides raw understanding regarding the feed selection and feed composition and can be helpful for refinery planning and scheduling of refinery assignments. Quality of products can be ensured for certain products whose design specifications were considered and rest could not be judged.

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NOMENCLATURE

$C_{e,d}$	=	cost of energy (Rs/day)
$C_{p,d}$	=	cost of products (Rs/day)
$C_{r,d}$	=	cost of raw materials (Rs/day)
C_c	=	cost of crude (Rs/bbl)
C_D	=	product value of diesel and atmospheric gas oil (Rs/lb)
C_e	=	cost of energy (Rs/kJ)
C_K	=	product value of kerosene (Rs/lb)
C_N	=	product value of naptha and heavy naptha (Rs/lb)
C_s	=	cost of steam (Rs/lb)
C_v	=	product value of light vacuum gas oil and heavy vacuum gas oil(Rs/lb)
F_A	=	flow rate of atmospheric gas oil (lb/hr)
F_D	=	flow rate of diesel (lb/hr)
F_{HN}	=	flow rate of heavy naptha (lb/hr)
F_H	=	flow rate of heavy vacuum gas oil (lb/hr)
F_K	=	flow rate of kerosene (lb/hr)
F_L	=	flow rate of light vacuum gas oil (lb/hr)
F_N	=	flow rate of naptha (lb/hr)
F_R	=	flow rate of residue (lb/hr)
$F_{a,s}$	=	flow rate of steam to ADU (lb/hr)
$F_{p,s}$	=	flow rate of steam to pre-flash tower (lb/hr)
$F_{s1,s}$	=	flow rate of steam to stripper 1(lb/hr)
$F_{s2,s}$	=	flow rate of steam to stripper 2 (lb/hr)
$F_{s3,s}$	=	flow rate of steam to stripper 3(lb/hr)
$F_{v,s}$	=	flow rate of steam to VDU (lb/hr)
$Q_{p,c}$	=	pre-flash condenser heat duty(kJ/sec)
$Q_{a,c}$	=	ADU condenser heat duty(kJ/sec)
$Q_{a,p1}$	=	ADU pump-1 heat duty(kJ/sec)
$Q_{a,p2}$	=	ADU pump-2 heat duty(kJ/sec)
$Q_{a,p3}$	=	ADU pump-3 heat duty(kJ/sec)
$Q_{v,p1}$	=	VDU pump-1 heat duty(kJ/sec)

$Q_{v,p2}$ = VDU pump-2 heat duty(kJ/sec)

$Q_{p,f}$ = pre-flash furnace heat duty(kJ/sec)

$Q_{a,f}$ = ADU furnace heat duty(kJ/sec)

$Q_{v,f}$ = VDU furnace heat duty(kJ/sec

1. INTRODUCTION

Crude oil distillation, considered as the most basic process of crude processing and petroleum refining.Generally, a crude oil distillationsystem (Fig. 1) refers to three parts a pre-flash tower, followed by an atmospheric distillation unit then followed by a vacuum distillation unit. The pre-flash tower is optional as per need. Since the feed used here consists of a larger proportion is lighter components so we have used the pre-flash tower to remove these lighter components and reduce the load on the following towers. The products of a crude distillation unit arebasically lightends,naphtha, diesel, kerosene, atmospheric gas oil, light and heavy vacuum gas oil and vacuum residue. An analysis of crude distillation is beneficial for having high process efficiency at low operating cost. Many a times, refineries involve blending of crudes due to operational and feed availability constraints, a feature that is more prominent in energy economy past globalization era.

Till now, many academicians have presented the analysis of crude distillation system. This work is based on thumb rule of basic grass-root design for CDU systems which involves simpler translation of crude assay data into suitable product distributions and then calculating the corresponding profit margins and trade-offs.This work provides raw understanding regarding the feed selection and feed composition and can be helpful for refinery planning and scheduling of refinery assignments during the crunch periods of higher demand less supply or lower demand high supplies. This work can also be helpful in choosing the feed based on the product requirements or profit margins required for operation.

The methodology adopted in this in this work involves design and optimization of a chosen CDU system configuration with steam flow constraints, design specification of products like naphtha, diesel and heat duties of pumps etc. with a specified number of plates, tray sizes in all the distillation columns using SQP optimization solver built in the software.

Several finer objectives of this article are addressed as follows:

i. Impact of crude selection on refinery profits and CDU optimality including feed and steam flow rates to the CDU system.

ii. Effect of binary crude compositions on refinery profit margins as well as CDU design variables such as steam and crude feed flow rates.

iii. Trade-offs associated to raw-materials and energy cost and profit margins of the CDU system for various choices of feeds.

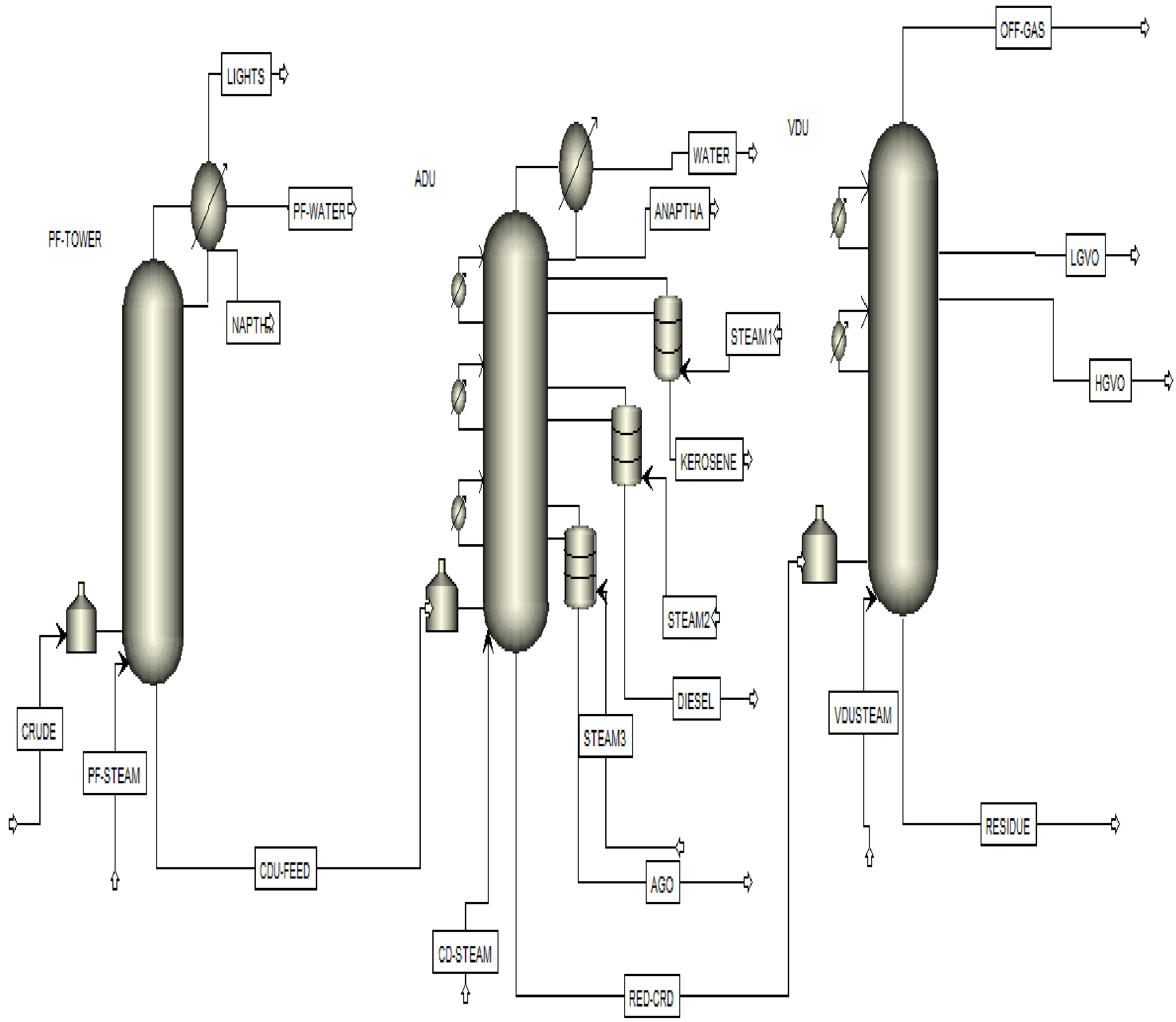

Fig1. Process flow diagram of CDU simulation

2. LITERATURE REVIEW

2.1. FLASH DISTILLATION:

It consists of vaporizing a definite fraction of liquid in such a way that the evolved vapour is in equilibrium with the residual liquid, separating the vapour from the liquid and condensing the vapour. Here are some equations that govern the distillation:

$$X_F = f\, Y_D + (1- f)X_B \qquad (1)$$

Where,

X_F	=	concentration of feed
f	=	molal fraction of feed that is vaporized and withdrawn continuously as vapour.
Y_D and X_B	=	concentration of vapour and liquid respectively.
$(1-f)$	=	molal fraction of feed that is vaporized and withdrawn continuously as liquid.

Another factor that is crucial in determining the ease of distillation is relative volatility. Relative volatility has to be either >>1 or <<1 for separation of components by distillation.

Mathematically,

$$\propto_{AB} = \frac{\frac{Y_{Ae}}{X_{Ae}}}{\frac{Y_{Be}}{X_{Be}}} \qquad (2)$$

Where,

$\propto_{AB}$=Relative volatility of two components (A and B)

Y_{Ae} = vapour concentration of component A

X_{Ae}= liquid concentration of component A

Y_{Be} = vapour concentration of component B

X_{Be} = liquid concentration of component B

In other ways this above equation can be written as:

$$Y = \frac{\propto X}{1+(\propto -1)X} \qquad (3)$$

Taking $X_B = (1 - X_A)$ and here Y and X can be understood as Y_{Ae} and X_{Ae}

The fraction f is not fixed directly but it depends upon the enthalpy of hot incoming liquid and the enthalpies of the vapour and liquid leaving the flash chamber.

$$H_F = f H_Y + (1- f)H_X \qquad (4)$$

Where,

H_F =enthalpies of the free liquid

H_Y = enthalpies of the vapor

H_X = enthalpies of the liquid product

MATERIAL BALANCES IN PLATE COLUMNS:

A distillation column (Fig.2) is fed with Fmol/hr of concentration $x_F$and delivers Dmol/hr of overhead product of concentrationx_D and B moles/hr of bottom product of concentration x_B. Two independent overall material balances can be written.

Total – material balance $F = D + B$ (5)

Component A balance $Fx_F = Dx_D + Bx_B$ (6)

Eliminating B from these equations gives$\frac{D}{F} = \frac{x_F - x_B}{x_D - x_B}$ (7)

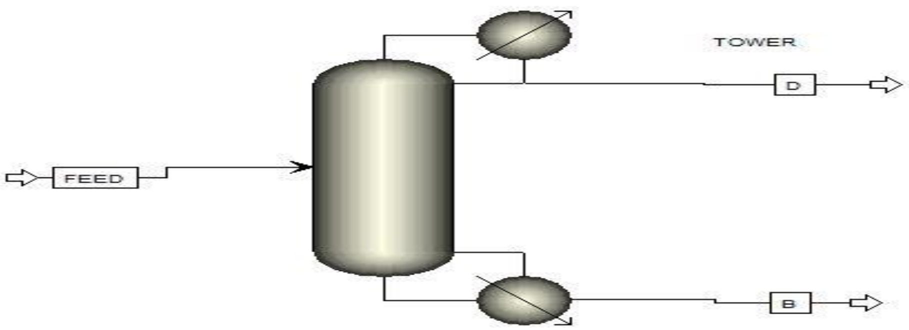

Fig2.Distillation column

REFLUX RATIO (R_D): It is the ratio of the reflux to the Overhead product

$$R_D = \frac{L}{D} = \frac{V-D}{D} \tag{8}$$

Operating line equation for distillation equation :

$$Y_{n+1} = \frac{R_D}{R_{D+1}} x_D + \frac{x_D}{R_D + 1} \tag{9}$$

MINIMUM REFLUX RATIO:

At any reflux less thantotal, the number of plates required for a given separation is larger than at total reflux and increases continuously as the reflux ratio is decreased. As the ratio becomes smaller, the number of plates becomes very large and at a definite minimum called the minimum reflux ratio.

MINIMUN NUMBER OF PLATES:

When V=L in the expression of reflux ratio, R_D is infinite, and we call this condition as *total reflux*. At this condition only we have minimum number of plates and also the rates of feed for the over-head and bottom product is zero. We have *Frenskee equation* which gives us the expression of minimum number of plates.

$$N_{min} = \frac{\ln[\frac{x_D\ (1-x_B\ )}{x_B(1-X_D)}]}{\ln\propto_{AB}} - 1 \tag{10}$$

PHASE LIQUID EQUILIBRIA IN MULTI- COMPONENT DISTILLATION

Vapour liquid equilibria of a mixture are described by distribution coefficients or K factors, and this K factor is described for each component and is the ratio of mole fractions in the vapour and liquid phases at equilibrium. Mathematically K is defined as:

$$K_i = \frac{y_{ie}}{x_{ie}} \tag{11}$$

DEW POINT AND BUBBLE POINT CALCULATION

The basic equation of BUBBLE POINT (initial boiling point of a liquid mixture)

$$\sum_{i=1}^{N_c} y_i = \sum_{i=1}^{N_c} K_i x_i = 1 \tag{12}$$

The basic equation of DEW POINT (initial condensation temperature)

$$\sum_{i=1}^{N_c} x_i = \sum_{i=1}^{N_C} \frac{y_i}{K_i} = 1 \qquad (13)$$

Where N_c is the number of components

2.2 CRUDE OIL PROPERTIES

The hydrocarbons in crude oil are mostly alkanes, cycloalkanes and various aromatic hydrocarbons while the other compounds contain nitrogen, oxygen and sulphur, and trace amounts of metals such as iron, nickel, copper and vanadium. The exact molecular composition varies widely from formation to formation but the proportion of chemical elements vary over fairly narrow limits as follows in Table 1. and Table 2

Table1. Composition by weight of different elements of crude oil

Element	Percentage
Carbon	83-85
Hydrogen	10-14
Nitrogen	.1-2
Oxygen	.05-1.5
Sulphur	.05-6
Metals	<.1

Table2. Composition by weight of different hydrocarbon constituents of crude oil

Hydrocarbon	Average	Range
Alkanes (paraffin)	30%	15 to 60%
Napthenes	49%	30 to 60 %
Aromatics	15%	3 to 30 %
Asphaltics	6%	remainder

2.3CRUDE OIL IMPURITIES

Produced crude oil contains many undesirable impurities. These impurities can present many varied problems during the refining process. The purpose of desalting is to remove these undesirable impurities from the crude oil stream prior to distillation.

The following are the most common types of impurities present in crude oil

- Inorganic Salts
- Crystalline and Dissolved solids
- Corrosion by-products such as Iron Oxide and Iron Sulphide
- Sand and Silt
- Drilling Mud and Polymers from Well Drilling/ Service works
- Oil Soluble Organic Chloride Compounds.

Inorganic salts

The most common inorganic salts present in the crude oil are

- Sodium Chloride, ($NaCl$)
- Magnesium Chloride, ($MgCl_2$)
- Calcium Chloride, ($CaCl_2$)

2.4. CDU PRODUCTS:

Flash tower:

- Naphtha
- Lights

CDU products and intermediates

- LPG
- Naphtha
- Kerosene
- Diesel
- Light gas oil
- Heavy gas oil
- Reduced crude oil

VDU products and intermediates

- Gas
- Light vacuum gas oil
- Heavy vacuum gas oil
- Vacuum residue

2.5. CRUDE DISTILLATION COLUMN

This column is provided with trays of which some trays are in the stripping section and some in the rectifying section. The crude distillation column flash zone Temperature is 365 ^{o}C and pressure is kept at 3.23 Kg/cm^{2}a respectively.

The crude feed enters the flash zone at tray. The vaporized portion of the feed along with the light ends from the stripping section is fractionated on trays above the flash zone to yield liquid side draw products, pump-around (circulating refluxes) and an overhead vapour stream. The various side streams taken out from Crude Distillation Column are: HGO, HGO-CR, LGO, LGO-CR, Kero and Kero-CR. RCO are obtained at the bottom of the column.

Crude Distillation Column: General Philosophy:

The major indicator for proper operation of Crude column is the column Temperature profile. Temperature is the variable most responsive to the changes in the operation and to the changes made by the operator. As along as the Crude quality remains constant & the heater outlet Temperature remains constant, the heat input to crude column remains constant. Therefore, the variables at the operator’s disposal are heat removal through Circulating reflux streams and product rates.

The Temperature of the side-cut is indicative of its endpoint at constant column pressure. The side-cut Temperature can be used as fairly an accurate means of holding or returning to an operating condition yielding a cut with the desired endpoint.

Product draw rate is used to control draw Temperature. If the product draw rate is increased, the end point of the side cut will rise. If the rate for one product is changed all the product qualities below will also change. If there is no desire to change the other products, the volume draw on the product below must be adjusted to balance the change in the product above.

The Circulating reflux stream flow rate is used to control the heat removed from the section in the column between the draw tray and the return tray. The primary aim is to adjust the reflux in the column to obtain correct separation between product streams. To improve separation, the vapor/liquid contact with in the column has to be increased. In order to improve vapor/liquid contact, the heat removal is reduced which increases the vapor flow up through the trays. The increased vapor flow results in a higher internal reflux down the trays. Reducing circulating reflux heat removal from the column will improve fractionation at the expense of high level heat recovery. Additional heat is removed further up the column at lower Temperature and ultimately in the overhead condensers.

In practice, once liquid and vapor rates have been established which provide adequate product separation, they do not have to be adjusted very often and the bulk of the changes in the column will relate to achieving volume yields, endpoint and flash point specifications.

3. **PROBLEM DESCRIPTION**

3.1 CRUDE ASSAY

Two crude assays namely BOMBAYHG (Bombay crude), ARABY (Arabian Light Crude) are considered in this work. Different binary fractions of these crude are taken as feed. Here, .Fig. 3a–b presents the true boiling point (TBP) curves of these crudes, respectively. As shown in the Table 3 naphtha range corresponds to 34.2 vol% in Bombay crude and 20 vol% in the Mideast crude respectively. A critical analysis of these crudes indicated that Bombay possesses more low boiling components whereas Araby crude possesses more high boiling components. Based on these TBP curves, the mixed crude TBP curves for various volume fractions of different single crudes have also been generated.

The curves Fig. 3 c-f presents the true boiling point (TBP) of various blends of both the crudes.

Table3.Crudes constituents based on TBP diagram

Components	Bombay-high crude	Araby crude(ARABYLT)
Naphtha	34.2%	20%
Kerosene	11%	12.2%
Diesel	14.3%	7.5%
AGO	22.6%	7.5%
LVGO	5.6%	7.5%
HVGO	18.8%	13.1%

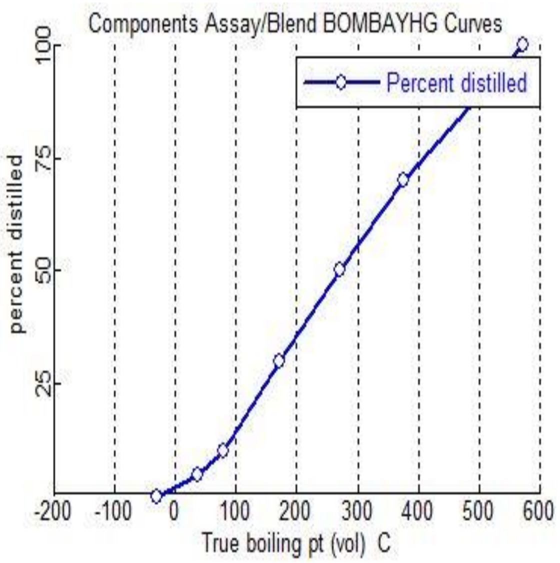

Fig3-a- percent distilled vs. TBP curve of Bombay High crude

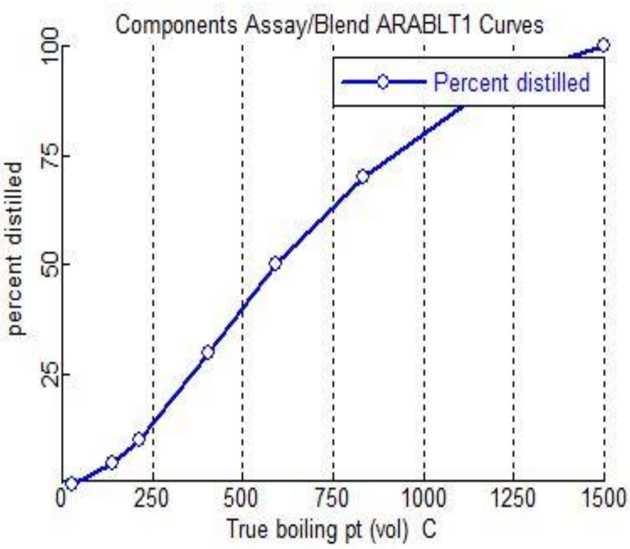

Fig 3-b- percent distilled vs. TBP curve of Arabian Light crude

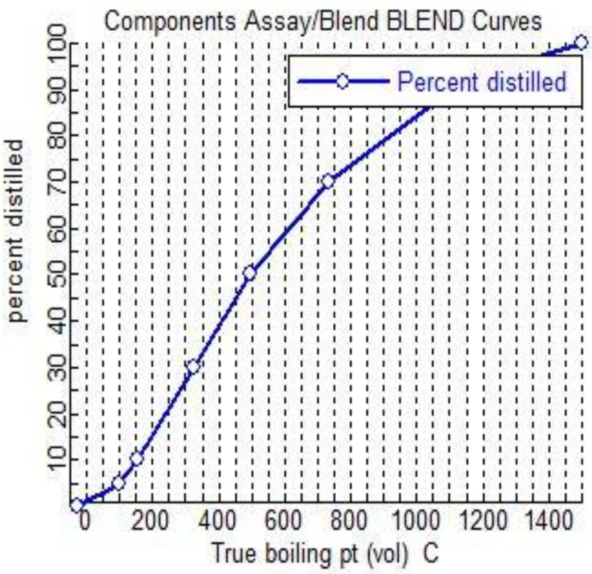

Fig3-c- percent distilled vs. TBP curve of BOMBAYHG and ARABLT in 20:80 ratio

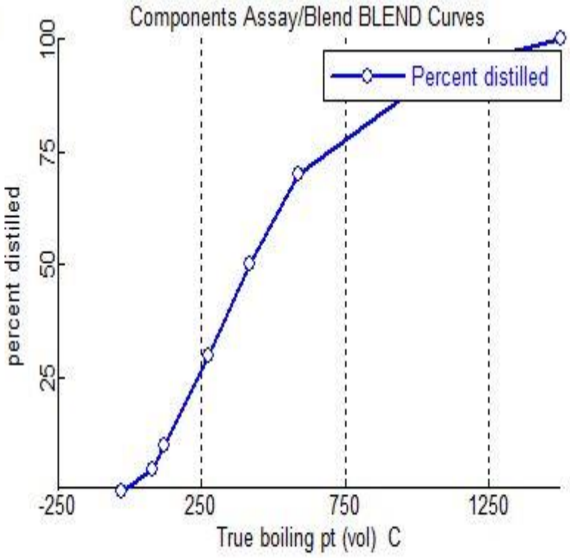

Fig 3-d- percent distilled vs. for blend of blend of BOMBAYHG and ARABLT in 40:60 ratio

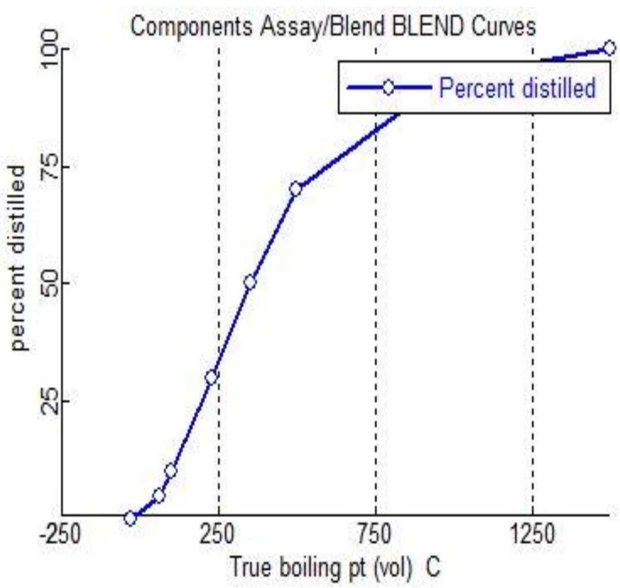

Fig 3-d- percent distilled vs. TBP curve of BOMBAYHG and ARABLT in 60:40 ratio

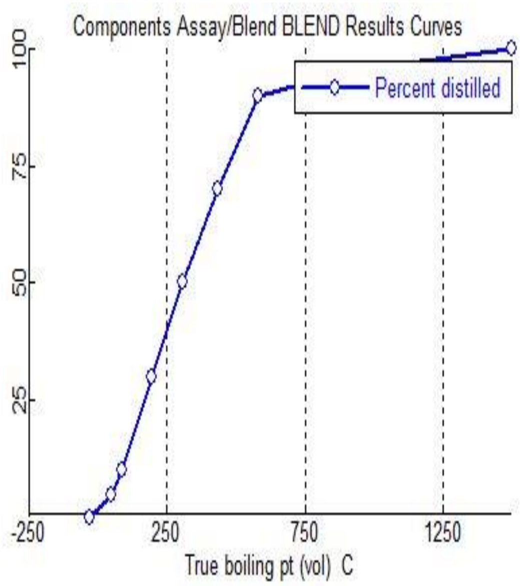

Fig 3-f- percent distilled vs. for blend of blend of BOMBAYHG and ARABLT in 80:20 ratio

3.2 PROCESS SETUP

The process flow sheet for the crude distillation unit is represented in the form of block diagram with the help of aspen plus commercial software. The representation of crude distillation unit consists of a pre-flash unit followed by an atmospheric distillation unit and a vacuum distillation unit. Feed streams to these streams have been subjected to undergo heating by the heating furnaces.

Here we have three furnaces for the three towers the bottoms of the towers act as the feed for the next tower. The bottoms of these towers are generally the high boiling components of these towers. Pump-around and side-strippers are present in the ADU and VDU columns. The columns are devoid of heaters as heating is generally done with the help of the steams provided. Side- stripping is also done with the help of steam.

Crude Assay (1) Name: ARABLT1
Geographic Source: Mideast Crude
Crude Assay Name: BOMBAYHG
Geographic Source: Asia/Pacific, former-Soviet Crude
Blend: .8 (std. vol fra) BOMBAYHG and .2 (std. vol fra) of ARABLT1
Process type: refinery
Base method: BK10

3.2.1 PRE-FLASH TOWER

The crude sent to the pre-flash furnace .undergoes partial vaporisation. The furnace operating Temp and pressure are taken to be 50 Pisa and 450 F. The furnace specification is chosen to be single stage flash at specified conditions.

The pre-flash tower contains 10 theoreticalstages. The condenser taken is a partial – vapour liquid condensate. No reboiler is used for this column. Valid phases for this tower are vapour, liquid and free water. The operating specification for this column is kept as – distillate flow rate with a standard volume of 15000 bbl/day.

Feed enters the stage 10 of the column through the furnace. Steam is introduced also at stage 10 with the convention of on-stage. Crude feed is removed from stage 10 with a liquid phase. Light ends are removed as vapour from stage 1. Naphtha is also removed as liquid from stage 1. Water is also removed as free-water from stage 1.

With the view from top to bottom, stage 1 pressure is taken to be 39.7 Pisa while, stage 2 pressure is taken to be 41.7 Pisa and bottom stage pressure is kept at 44.7 Pisa. Type of condenser is chosen to be partial vapour-liquid condenser with a temperature of 197 F.

We have to specify the product quality of naphtha using ASTM 95% Temperature and manipulate the distillate rate toachieve the target

The naphtha product design specification

Type: ASTM D86 Temperature (dry,liquid volume basis)

Target- 345 F

Liquid- 95%

3.2.2 ATMOSPHERIC DISTILLATION UNIT (ADU)

The bottom product of the pre-flash tower enters the process furnace and undergoes a rapid vaporisation of about3% by volume. The vaporisation is a fractional over flash and with a standard volume of .03 and the furnace pressure is maintained at 24.18 Pisa. The ADU column has 25 stages (chosen from earlier studies) and consists of a total condenser. There is no reboiler in the tower as steam is provided at the bottom. The steam temperature and pressure used in this column is 60 Pisa.

The ADU system is coupled with 3 side-strippers and 2 pump-around circuits. The feed enters the system at the 22nd stage of the column and the convention is at the furnace. Steam(CD-STEAM) for the tower enters the tower at the 25th stage and the convention is On-stage. The pressures inside the column for stage 1, stage2 and bottom stage are 15.7 Pisa, 20.7 Pisa and 24.7 Pisa respectively.

Three side strippers are used. For stripper 1, number of stages are 4, stripper product is kerosene, liquid draw is from 6th stage and overhead return is at the 5thstage. Stripping steam is CD-STM1. Bottom product flow is 11050 bbl/day.

For stripper 2, number of stages are 3, stripper product is diesel, liquid draw is from 13th stage and overhead return is at the 12thstage. Stripping steam is CD-STM2. Bottom product flow is 16500 bbl/day.

For stripper 3, number of stages are 2, stripper product is AGO, liquid draw is from 18th stage and overhead return is at the 17thstage. Here stripping steam is CD-STM3. Bottom product flow is 7500 bbl/day.

Two pump-around circuits are used to facilitate internal reflux.For first around (P-1), draw stage is 8 and return stage is 6, draw off type: partial, standard flow is 49000 bbl/day with a heat duty of -40 MMBtu/hr. For second pump around (P-2), draw stage is 14 and return stage is 13, draw off type: partial, standard flow is 11000 bbl/day with a heat duty of -15 MMBtu/hr.

Design specifications: 1st specification is for HNAPTHA(naphtha stream), Design type: ASTM D86(dry volume basis). The target temperature is 464 F with a liquid % of 95.

For achieving this specification parameterspecified that is to be achieved is Distillate Flow Rate.

2nd specification is for DIESEL, design type is ASTM D86 (dry volume basis) target tempto be achieved is 640 F and the liquid % to be achieved is 95%.

Table4.Mass flow rate of steam used in ADU column

Streams(steam)	Mass flow rate
CD-STEAM(bottom of tower)	13000
CD-STM1 (stage 6)	3300
CD-STM2 (stage 13)	1000
CD-STM3 (stage 18)	800

3.2.3 VACUUM DISTILLATION UNIT (VDU)

The bottom product of the atmospheric distillation unit is fed to the vacuum distillation unit which is facilitated with jet ejectors that enable the generation of vacuum in the unit. The vacuum unit separates the atmospheric column bottom product into off-gas, light vacuum gas oil (LVGO), heavy vacuum gas oil (HVGO) and residual oil.

The vacuum distillation consists of 6 theoreticalstages with the pressure maintained at 60 mm Hg at the top and 70 mm Hg at the bottom. Configuration of the condenser is None-top pump around and that of re-boiler is none- bottom feed.

Red-crude /bottom enters at the furnace i.e. stage 6. VDU-STM enters at stage 6.

In the case of products, residue is extracted out from stage 6 as liquid phase. LGVO is extracted as total liquid from stage 2 with a flow rate of 8000 bbl /day. HGVO is extracted as liquid from stage 4 with a flow rate of 17000 bbl/day. Off-gas is removed as vapour from stage 1.

Furnace specification is: the type of furnace used here is single stage flash with liquid run back. The fractional over flash is of standard volume of 0.006. The furnace pressureis 2.03psia.

Two pump-around circuits are used here, for first pump (P-1) draw stage is 2 and return stage is 1 with a draw-off type aspartial. Thestandard flow maintained is 20000 bbl /day a heat duty is –28.6 MMBtu/hr.

For first pump (P-2) draw stage is 4 and return stage is 3 draw off type is partial having a standard flow is 49000 bbl. /day and a heat duty is -80 MMBtu/hr.

Design specification:

Specification type: stage temperature of stage 1, whose temp to be achieved at that stage is 150F. The variable parameter chosen is pump around duty of pump 1.

3.3 COST ANALYSIS:

Since profit making remains to be the sole purpose behind every setup, so the objective function is profit maximisation,which is calculated using the feed cost, capital and energy cost and products costs. Equations 14, 15, 16 & 17 are to be used to calculate the objective function. Cost prices required for calculating the objective function are shown in Table.5 and flow rates are given in Fig.4, Fig.5 and Fig.6.

$$C_{r,d} = C_c F_c + C_s (F_{p,s} + F_{a,s} + F_{s1,s} + F_{s2,s} + F_{s3,s} + F_{v,s}) \tag{14}$$

$$C_{e,d} = C_e(Q_{p,c} + Q_{a,c} ++ Q_{a,p1} + Q_{a,p2} + Q_{a,p3} + Q_{v,p1} + Q_{v,p2} + Q_{p,f} + Q_{a,f} + Q_{v,f}) \tag{15}$$

$$C_{p,d} = C_N(F_N + F_{HN}) + C_K F_K + C_D(F_D + F_A) + C_V(F_L + F_H) + C_R F_R \tag{16}$$

Ojective functon for cost analysis

$$O_m = C_{p,d} - C_{e,d} - C_{r,d} \tag{17}$$

Table 5. Unit prices of various commodities (2012)

Unit prices of various commodities		
commodities	Price	Units
Crude oil(Bombay High crude) (C_C)	6007	Rs/bbl
Crude oil(Arabian Light crude) (C_S)	7000	Rs/bbl
Naphtha (C_N)	28.61	Rs/lb
Kerosene (C_K)	28.55	Rs/lb
Vacuum gas oil (C_V)	23.9	Rs/lb
Residue (C_R)	18.95	Rs/lb
Diesel (C_D)	30.02	Rs/lb
Steam (C_S)	.332	Rs/lb
Energy (Ce)	2.844×10^{-4}	Rs/KJ

4. RESULTS AND DISCUSSION

On simulation of the model as per the given parameters and conditions the following results are obtained.

Display: Streams Format: PETRO_E Stream Table

	FEED	PF-STEAM	CRD-FEED	LIGHTS	NAPTHA	PF-WATER	
Temperature F	200.0	450.0	498.7	197.0	197.0	197.0	
Pressure psi	60.0	60.0	44.7	39.7	39.7	39.7	
Mass Flow lb/hr	1225353.9	3068.0	801942.7	17245.9	399068.7	10164.6	
Enthalpy MMBtu/hr	-999.2	-17.2	-470.5	-21.2	-317.4	-68.1	
Vapor Frac	0.000	1.000	0.000	1.000	0.000	0.000	
Average MW	173.9	18.0	302.0	54.7	108.4	18.0	
Liq Vol 60F bbl/day							
WATER	656.3	210.5	4.0	105.1	60.4	697.3	
PC59F	5908.1		84.0	1085.8	4738.3		
PC138F	1515.2		48.6	104.7	1361.9		
PC163F	1843.3		75.2	89.1	1679.0		
PC188F	2354.3		124.3	77.4	2152.6		

Fig 4.Stream results of pre-flash column

Display: Streams Format: PETRO_E Stream Table

	CRD-FEED	CD-STEAM	CD-STM2	CD-STM3	CD-STM1	RED-CRD	KEROSENE	DIESEL	AGO	HNAPTHA	CD-WATER
Temperature F	498.7	450.0	450.0	450.0	450.0	638.6	426.3	516.5	593.5	193.6	193.6
Pressure psi	44.7	60.0	60.0	60.0	60.0	24.7	21.2	22.4	23.3	15.7	15.7
Mass Flow lb/hr	801942.7	8396.7	703.6	612.9	1896.8	474337.4	135465.1	29261.2	90219.9	72765.4	11503.7
Enthalpy MMBtu/hr	-470.5	-47.1	-3.9	-3.4	-10.6	-235.0	-83.8	-17.7	-50.3	-54.7	-77.1
Vapor Frac	0.000	1.000	1.000	1.000	1.000	0.000	0.000	0.000	0.000	0.000	0.000
Average MW	302.0	18.0	18.0	18.0	18.0	434.9	210.2	255.8	307.7	140.2	18.0
Liq Vol 60F bbl/day											
WATER	4.0	576.0	48.3	42.0	130.1	0.3	3.5	0.1	< 0.1	7.4	789.2
PC59F	84.0					< 0.1	< 0.1	< 0.1	< 0.1	84.0	
PC138F	48.6					< 0.1	< 0.1	< 0.1	< 0.1	48.6	
PC163F	75.2					< 0.1	< 0.1	< 0.1	< 0.1	75.2	
PC188F	124.3					< 0.1	0.1	< 0.1	< 0.1	124.2	

Fig 5- stream result of ADU column

Display: Streams | Format: PETRO_E | Stream Table

	RED-CRD	VDU-STM	RESIDUE	LGVO	HGVO	OFF-GAS
Temperature F	638.6	450.0	737.5	401.5	667.4	150.0
Pressure psi	24.7	60.0	1.4	1.2	1.3	1.2
Mass Flow lb/hr	474337.4	11471.8	29659.6	204329.0	240265.9	11554.7
Enthalpy MMBtu/hr	-235.0	-64.3	-13.1	-139.0	-110.3	-65.9
Vapor Frac	0.000	1.000	0.000	0.000	0.000	1.000
Average MW	434.9	18.0	735.7	378.2	471.5	18.1
Liq Vol 60F						
WATER	0.3	787.0	< 0.1	0.2	< 0.1	787.1
PC59F	< 0.1		< 0.1	< 0.1	< 0.1	< 0.1
PC138F	< 0.1		< 0.1	< 0.1	< 0.1	< 0.1
PC163F	< 0.1		< 0.1	< 0.1	< 0.1	< 0.1
PC188F	< 0.1		< 0.1	< 0.1	< 0.1	< 0.1

Fig 6- stream result of VDU column

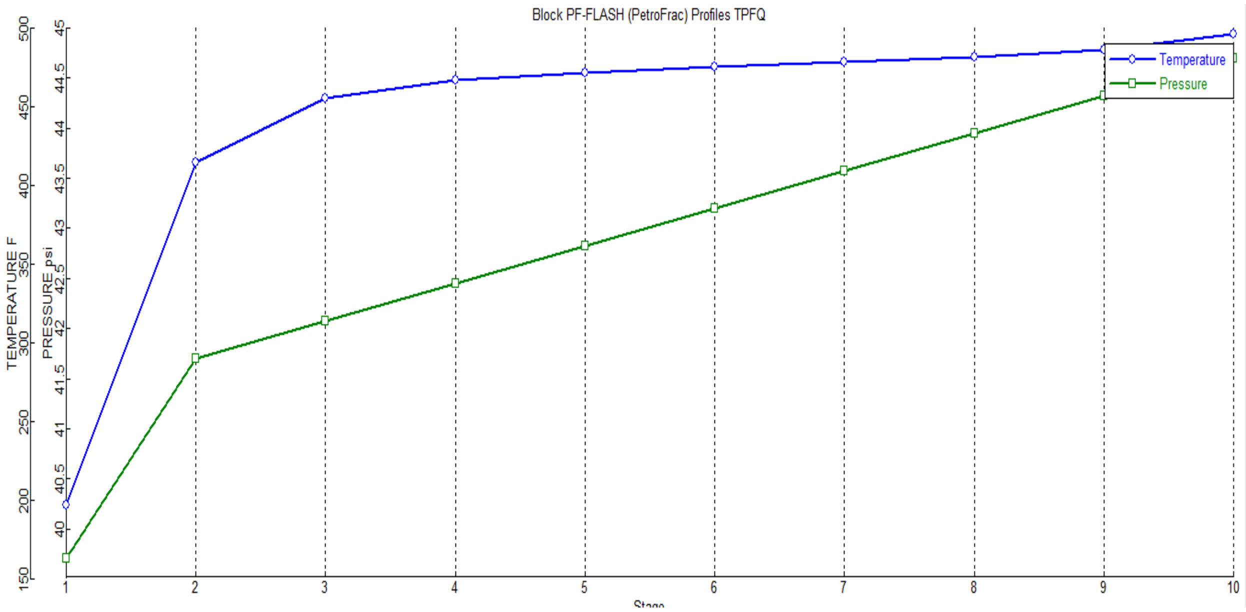

Fig 7: Temp and pressure vs. stage of pre-flash tower

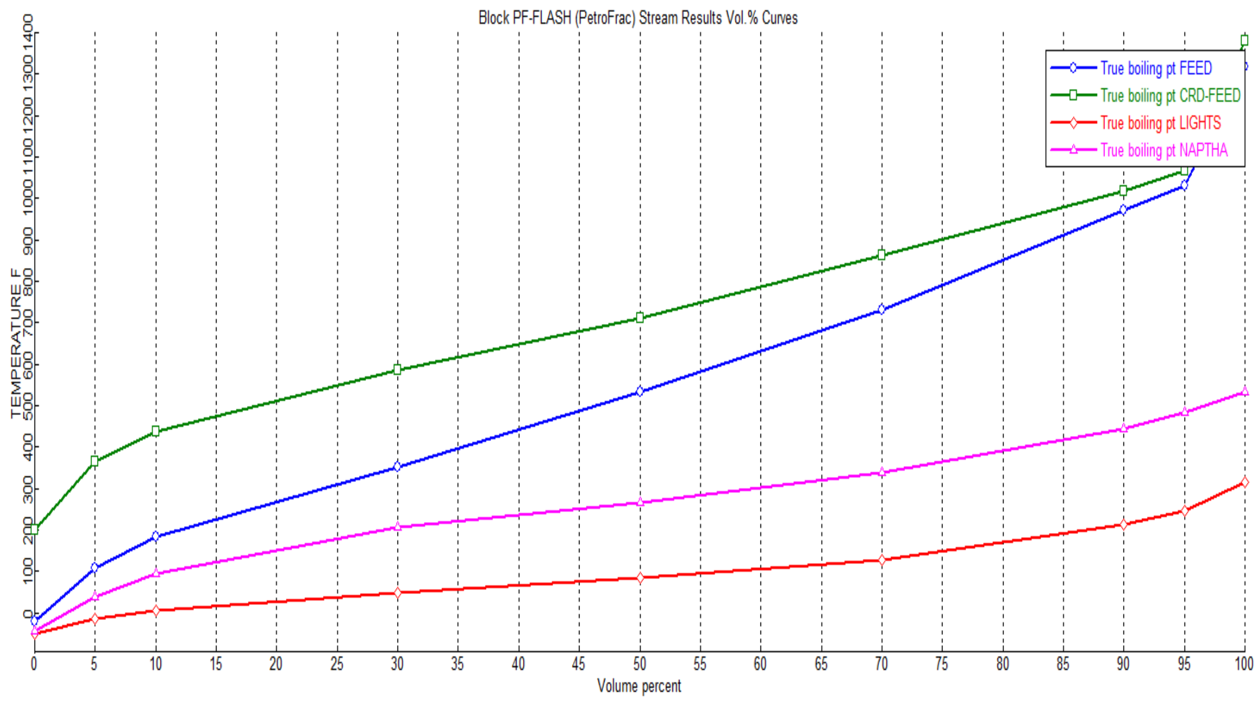

Fig 8: TBP of all streams vs volume percent of pre-flash tower

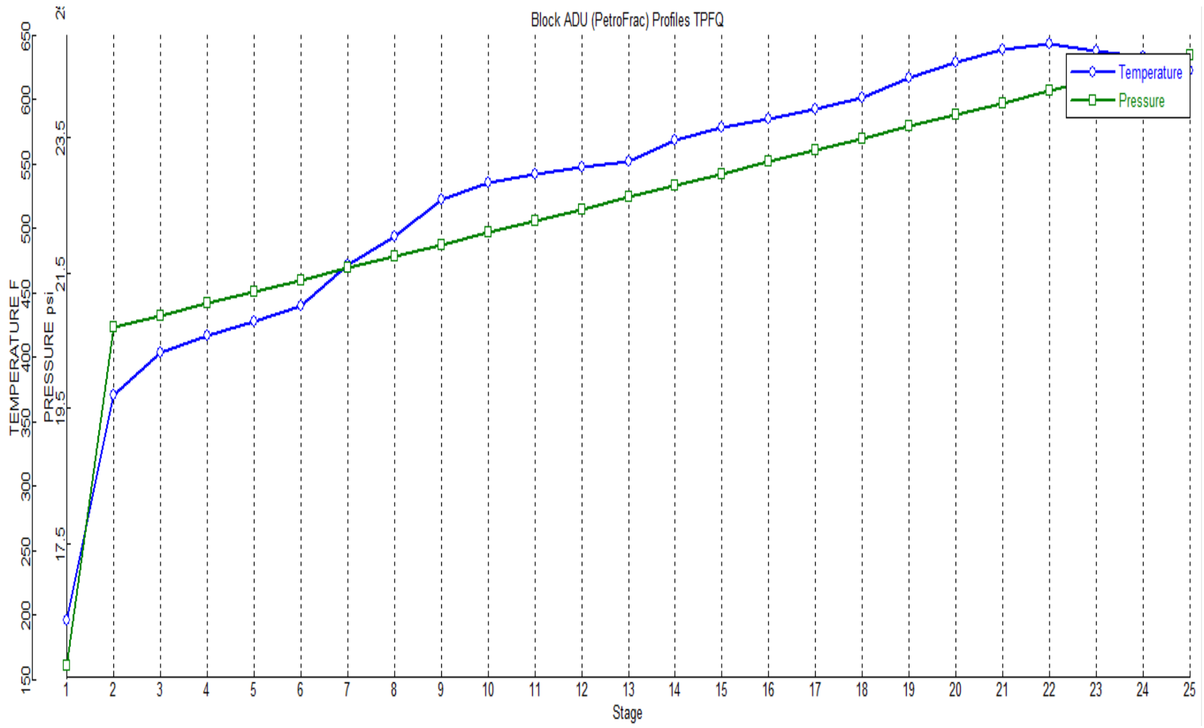

Fig 9: Temp and pressure vs. stage of ADU column

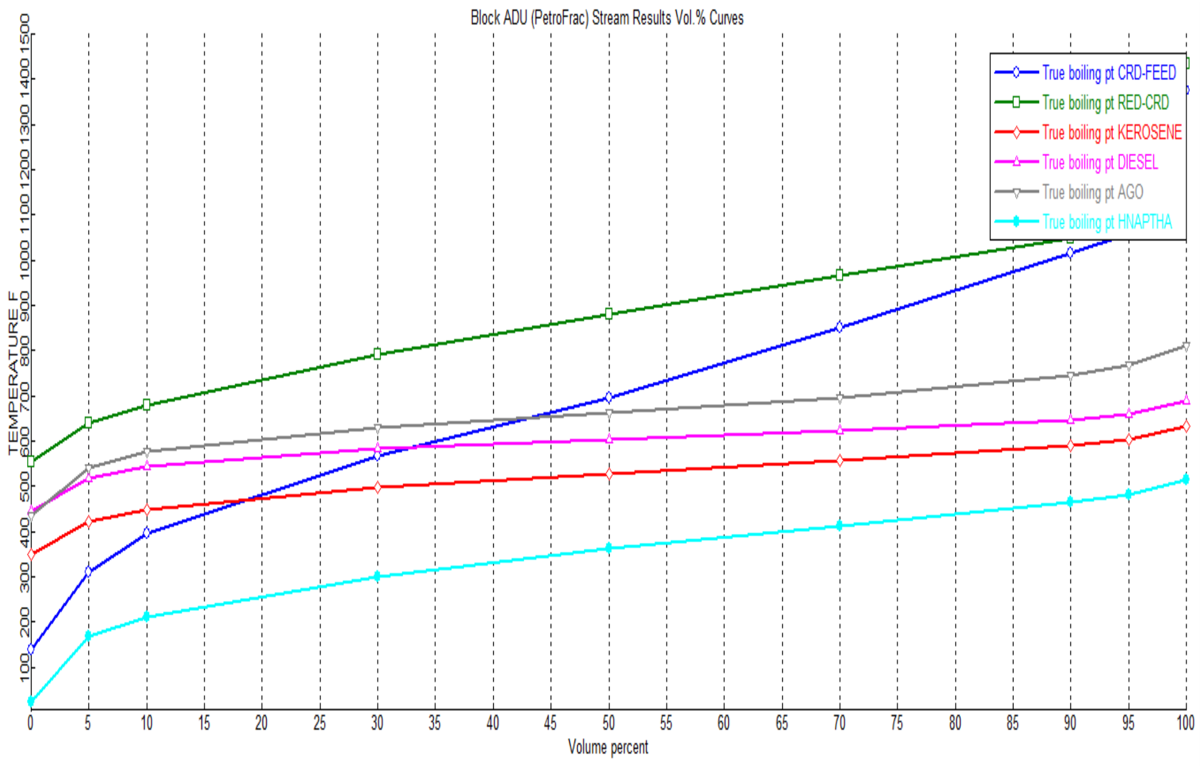

Fig 10: TBP of all streams vs volume percent of ADU column

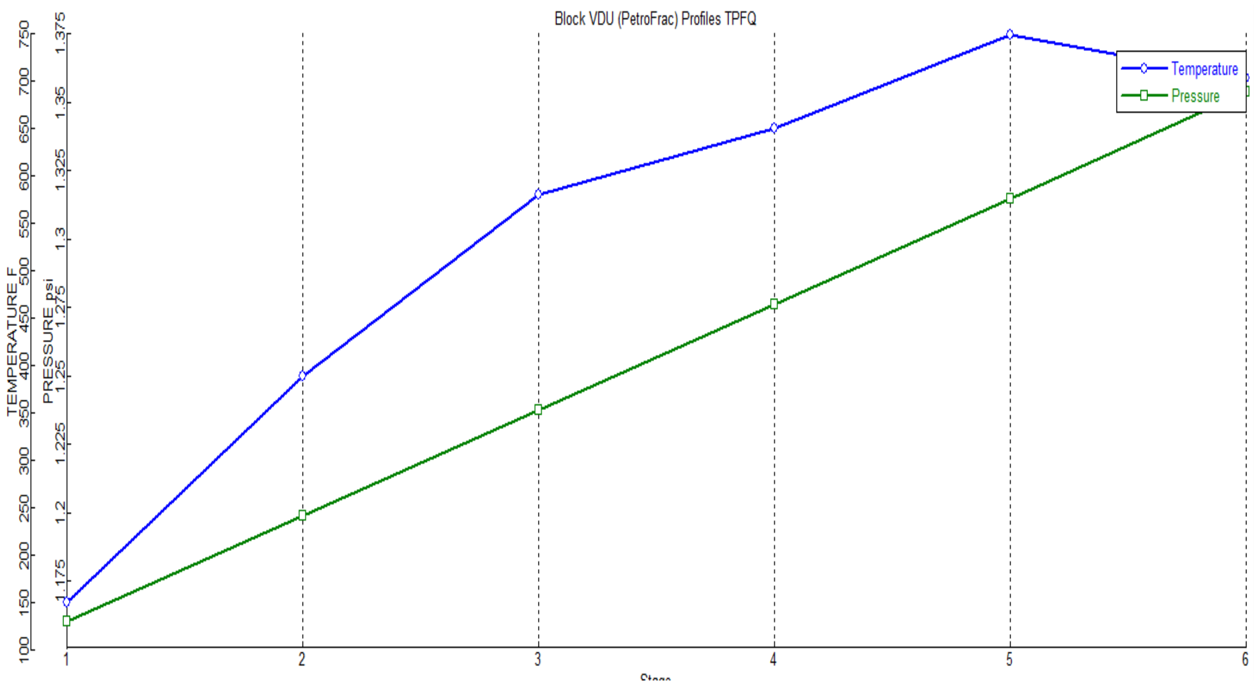

Fig 11: Temp and pressure vs. stage of VDU column

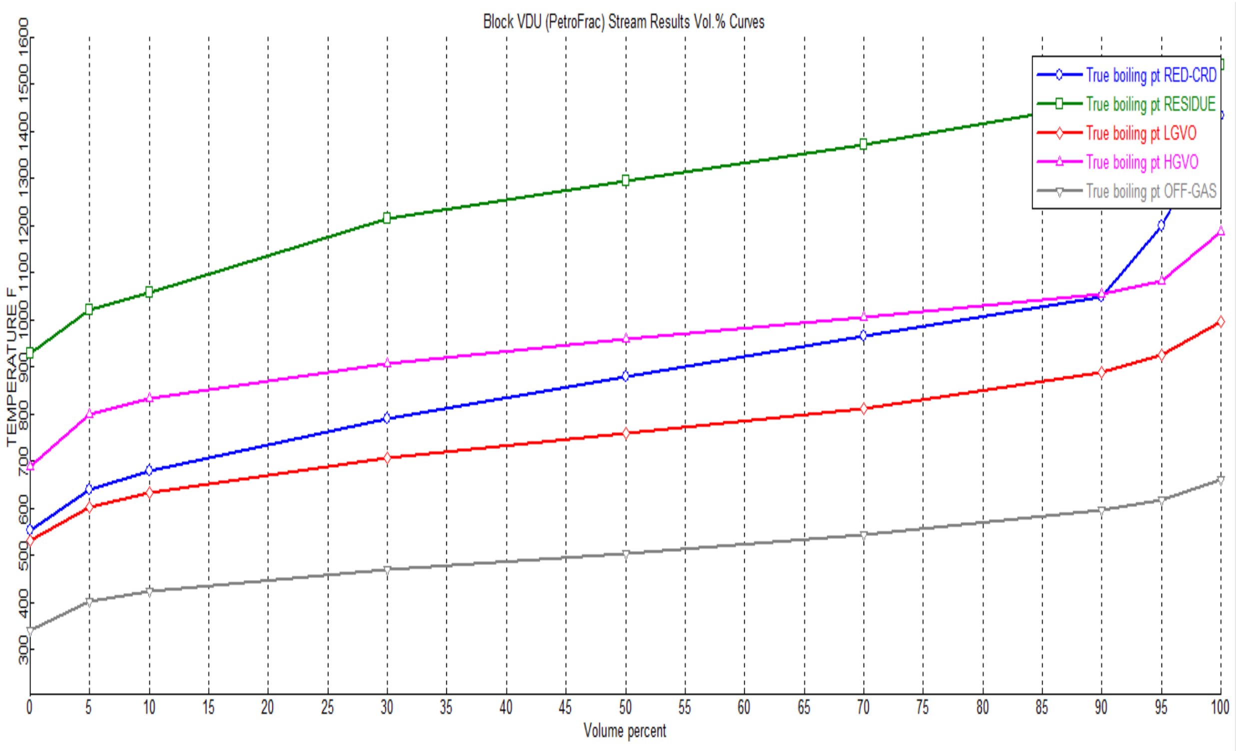

Fig 12: TBP of all streams vs. volume percent of VDU column

DISCUSSION

Given below in Table 6, product flow rates of different feeds are shown. Fig13, Fig14, Fig15, Fig16 comparison of the product flow rates has been diagrammatically shown.

Table 6. Product flow rates of different blends of feed

Streams/blend ratio->	80:20	60:40	40:60	20:80
AGO	89842.1	91145.4	92522.1	93973
Diesel	77134.7	72873.9	68531.8	62083.8
HGVO	240546.5	239813.6	238991.4	238229.2
HNAPTHA	156361.3	157683.5	159274.5	161257.6
Kerosene	135031.3	134511.1	133996.6	133484
LGVO	190072.3	186500.9	181391.7	175875
Lights	47331.7	34573.8	22029.6	11058
Naphtha	256500	249985.1	243076.3	234446.5
Red-crude	458613.7	487625.1	516584	545494.8
Residue	27895.6	61218.2	96115.6	131312.3

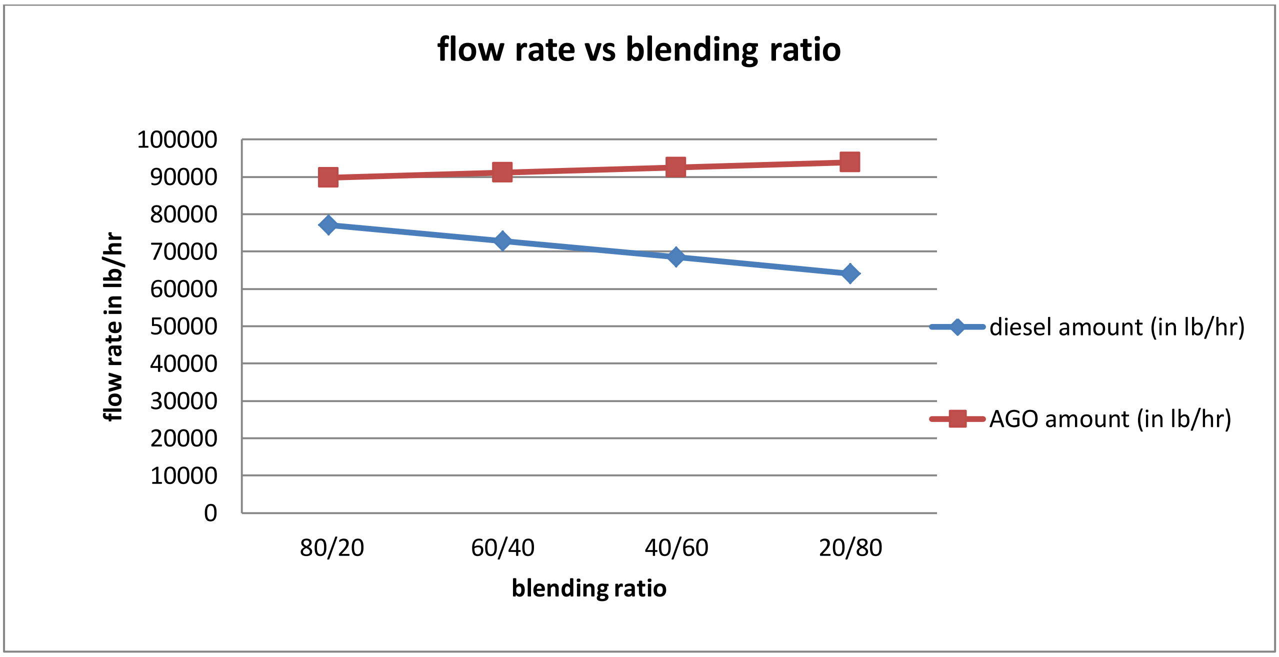

Fig 13: Comparison of flow rates of Diesel and Atmospheric gas oil (AGO) of different feed blends

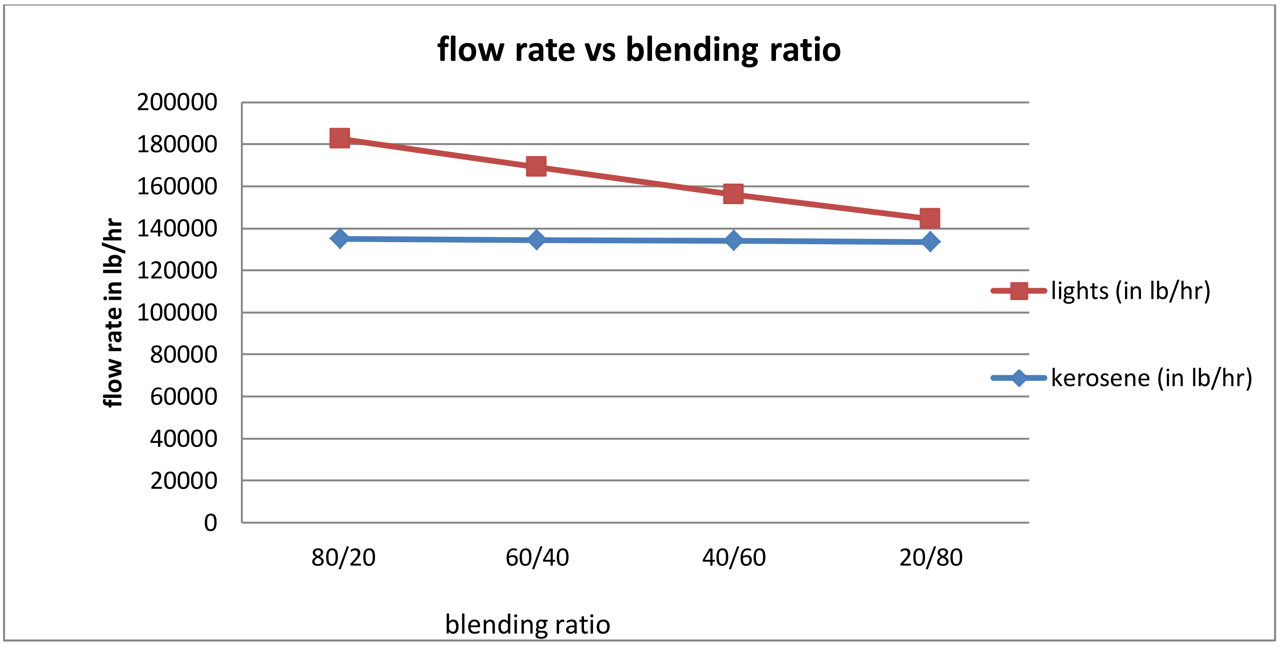

Fig 14: Comparison of flow rates of Kerosene and lights of different feed blends

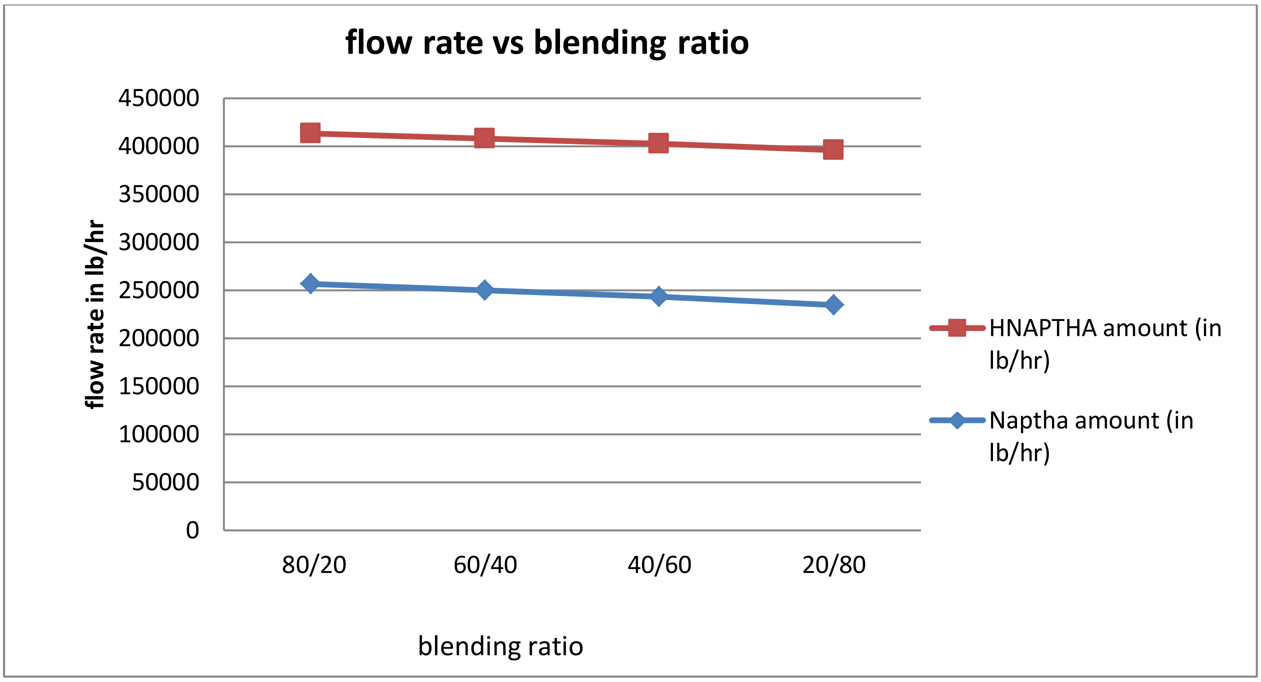

Fig 15: Comparison of flow rates of naphtha and HNAPTHA of different feed blends

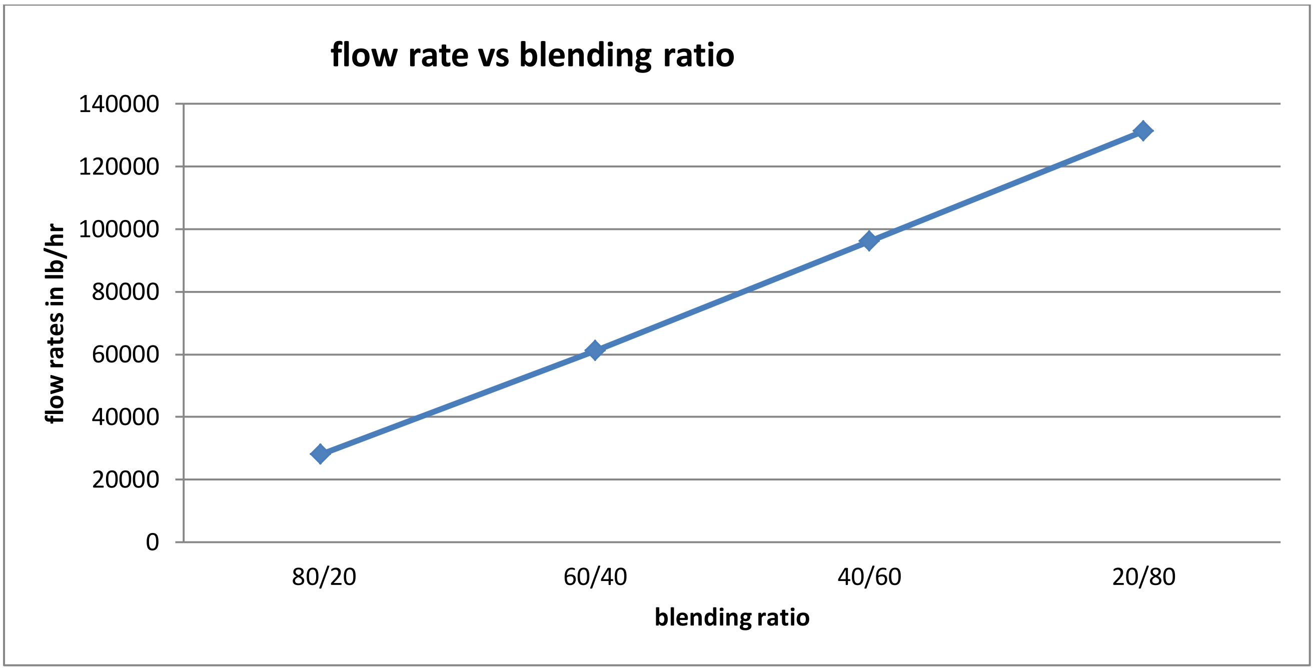

Fig 16: Comparison of flow rate of residue of different feed blends

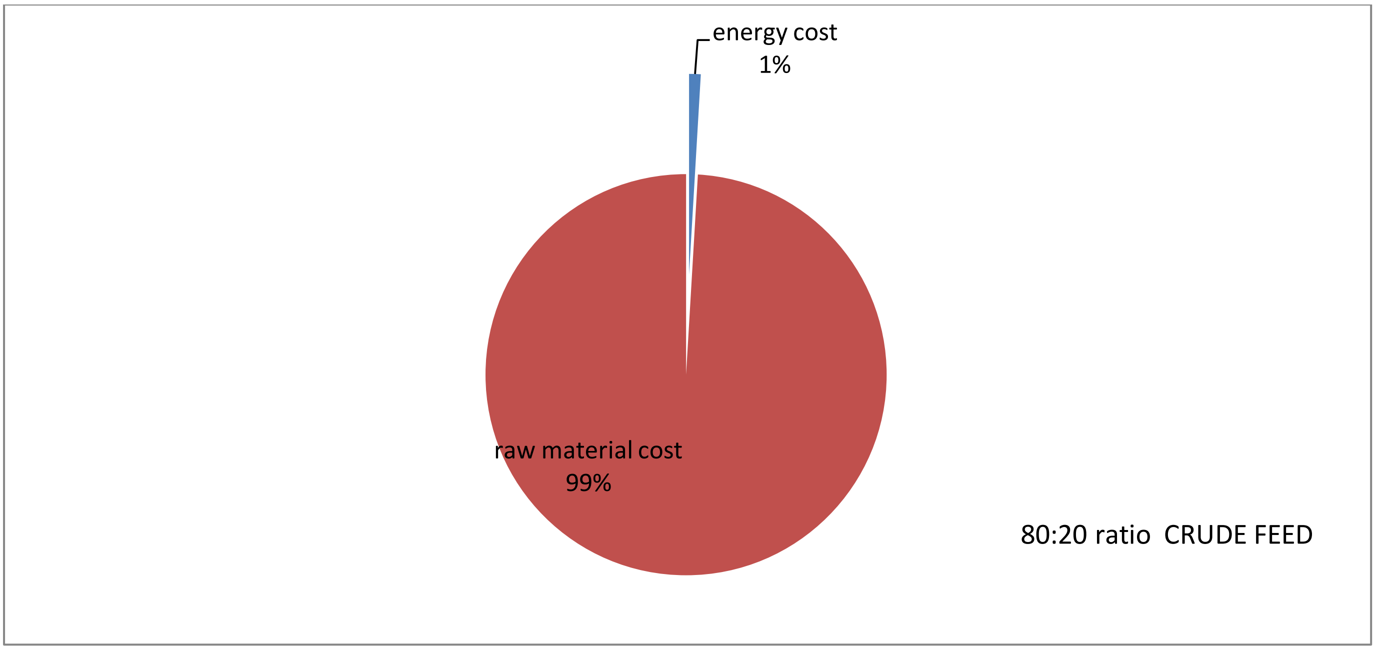

Fig 17: Comparison of cost of raw materials vs. Cost of energy of 80:20 ratio crude feed blend

The objective function defined earlier in the text has been calculated and shown in Fig18 .This figure (Fig18) gives us the clear idea about the Total profit, Total feed cost, and Total income per day made.

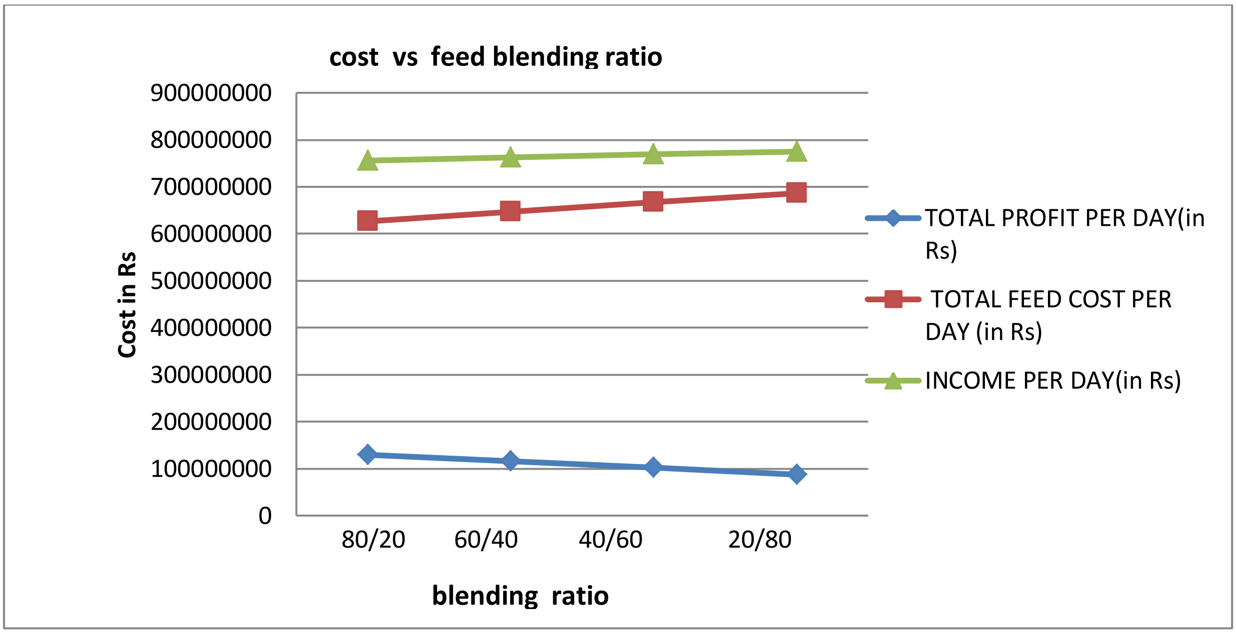

Fig 18: Comparison of feed cost, income, and profit with different blends of feed

5. CONCLUSION

This work presents the results of choosing a crude feed on the product amount, its quality and the profit margins obtained, which in turn can help in refinery planning and scheduling. This work has considered a grass root design of crude distillation column consisting of a pre-flash, vacuum distillation and atmospheric distillation towers with minimal number of stages/plates, pump around and strippers of the columns .As always, the performance of any plant is analysed on the profits made by the plant, so here the objective function considered is maximisation of profits made per day.

Inferences drawn from above works:

a) Crude blend of 80:20 ratio of Bombay high and Arabian Light provided the highest profit and crude blend of 20:80 ratio of Bombay high and Arabian Light provided the lowest profit.
b) Residue amount is highest in crude blend of 20:80 ratio of Bombay high and Arabian Light, so the need arises for cracking, reforming units in the downstream of plant process for obtaining more amounts of products.
c) Naphtha and light ends amount is found highest in crude blend of 80:20 ratio of Bombay high and Arabian Light.
d) Raw material costs dominated the energy costs as this contributed only 1% of overall input costs.
e) With minimal operating cost and with maximum profit crude blend of 80:20 ratio of Bombay high and Arabian Light is the finest crude according to the objective function.

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